REMARKS

In the outstanding Office Action, the restriction requirement was confirmed. Through this response, the non-elected claims have been canceled.

Claims 16-79 were allowed and Claims 5-10, 12, 14 and 15 were objected to as depending from a rejected base claim, but were also otherwise indicated as allowable.

Claims 1-4, 11 and 13 were rejected only on the basis of obviousness over *Ilmarinen et al.*, United States Patent No. 5,495,678 on the basis of calculations based on the data in the '678 specification.

The calculations relied upon do not reflect flow parameters in the sheet of the '678 patent nor do they render the subject mater of Claim 1 obvious. Claim 1 has been amended to make most clear that the flow parameters in the sheet must be used to calculate throughdrying coefficient and Reynolds Number.

The calculations in the Office Action rely on a faulty assumption that G, the mass flow, rate can be estimated as ΔP/V. It cannot, as is seen from the relationship between pressure drop and velocity on page 20 of the application as filed:

$$-\frac{dP}{dx} = \alpha\mu V / g_c + \beta\rho V^2 / g_c \qquad [4]$$

The Examiner's attention is also directed to Exhibits 1 and 2 hereto (excerpts from Kirk-Othmer and Perry's) showing similar relationships.

The faulty assumption in the Office Action as to G leads to the calculation result that mass flow rate **decreases** with velocity; a clear error, since the mass flow rate is related to the velocity via the relationship G = ρV (page 21, line 6 of the application as filed). In any event, one calculates the Reynolds Number and throughdrying coefficient as is set forth on pages 29-31 of the application as filed. The '678 reference does not provide enough information to make such calculations even if one knew the hydraulic properties of the sheet disclosed therein (which are also not provided) because only jet velocity, not bulk velocity is given.

The '678 patent is directed to throughdried (TAD) sheet. Applicants have simulated the Reynolds Numbers for that Material in Table 2 of the application as filed, where it can be seen that the Reynolds Numbers and throughdrying coefficients are outside the ranges of Claim 1, which is accordingly believed allowable. The values obtained for TAD material appear on page 37 of the application as filed and are repeated below:

Table 2 Excerpt–
Hydraulic Diameter, Void Volume Fraction, and Throughdrying Coefficient for TAD Sheet

Example	Category	Hydraulic Diameter	Reynolds Number	Void Volume Fraction	Through-Drying Coefficient
A	Simulated TAD	1.704E-05	1.500	0.771	3.333
B	Simulated TAD	1.382E-05	2.036	0.803	2.982
C	Simulated TAD	8.324E-06	1.144	0.799	3.749
D	Simulated TAD	1.330E-05	2.111	0.820	2.947
E	Simulated TAD	3.889E-05	11.952	0.814	2.167
F	Simulated TAD	3.871E-05	13.327	0.811	2.150
G	Simulated TAD	2.858E-05	9.549	0.826	2.209
H	Simulated TAD	1.267E-05	4.876	0.846	2.410
I	Simulated TAD	1.255E-04	48.211	0.835	2.041
J	Simulated TAD	4.534E-05	16.162	0.821	2.124
K	Simulated TAD	1.372E-05	5.888	0.836	2.340
L	Simulated TAD	3.320E-05	11.368	0.812	2.176

All claims should be allowed.

Respectfully submitted,

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December 2, 2003

Exhibit 1

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Kroschwitz

EDITOR
owe-Grant

KIRK-OTHMER

ENCYCLOPEDIA OF CHEMICAL TECHNOLOGY

FOURTH EDITION

VOLUME 11

FLAVOR CHARACTERIZATION
TO
FUEL CELLS

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of turbulence is generated at the jump because of the sudden lateral expansion, and energy is dissipated through friction.

Flow in Porous Media. Flow of fluids through fixed beds of solids occurs in situations as diverse as oil-field reservoirs, catalyst beds and filters, and absorption (qv) towers. The complex interconnected pore structure of such systems makes it necessary to use simplified models to make practical quantitative predictions. One of the more successful treatments of single-phase pressure drop through such systems employs the results for flow through tubes, using average velocities and tube diameters. The average velocity through the pores is related to the superficial velocity, ie, the apparent velocity based on the entire cross section, via

$$\overline{V} = \frac{V_s}{\epsilon} \tag{19}$$

For spherical particles the average diameter of the pores, defined as four times the pore volume divided by the surface area, can be shown to be

$$\overline{D} = \frac{6\epsilon}{1-\epsilon} d_p \tag{20}$$

The overall pressure drop is expressed as the sum of a laminar term proportional to $\overline{V}/\overline{D}^2$ and a turbulent term proportional to $\overline{V}^2/\overline{D}$ to yield the Ergun equation (1):

$$\frac{\Delta P}{L} = A \frac{\mu V_s}{d_p^2} \frac{(1-\epsilon)^2}{\epsilon^3} + B \frac{\rho V_s^2}{d_p} \frac{(1-\epsilon)}{\epsilon^3} \tag{21}$$

For randomly packed spherical particles, the constants A and B have been determined experimentally to be 150 and 1.75, respectively. For nonspherical particles, equivalent spherical diameters are employed and additional corrections for shape are introduced.

When two phases are present the situation is quite complex, especially in beds of fine solids where interfacial forces can be significant. In coarse beds, eg, packed towers, the effects are often correlated empirically in terms of pressure drops for the single phases taken individually.

Non-Newtonian Fluids: Die Swell and Melt Fracture. For many fluids the Newtonian constitutive relation involving only a single, constant viscosity is inapplicable. Either stress depends in a more complex way on strain, or variables other than the instantaneous rate of strain must be taken into account. Such fluids are known collectively as non-Newtonian and are usually subdivided further on the basis of behavior in simple shear flow, ie, flow between sliding planes or, to a good approximation, between two mutually rotating cylinders. Figure 10 illustrates the behavior of several of these. The types illustrated are examples of the generalized Newtonian fluid. This simple generalization uses the Navier-Stokes equations but takes viscosity to depend in some way on a single quantity: the rate of strain.

Exhibit 2

PERRY'S CHEMICAL ENGINEERS' HANDBOOK SEVENTH EDITION

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tube bundle, gas properties should be evaluated at the mean temperature

$$T_m = T_t + K\,\Delta T_{lm} \tag{6-162}$$

where T_t = average tube-wall temperature
K = constant
ΔT_{lm} = log-mean temperature difference between the gas and the tubes.

Values of K averaged from the recommendations of Chilton and Genereaux (*Trans. AIChE*, **29**, 151–173 [1933]) and Grimison (*Trans ASME*, **59**, 583–594 [1937]) are as follows: for in-line tubes, 0.9 for cooling and −0.9 for heating; for staggered tubes, 0.75 for cooling and −0.8 for heating.

For nonisothermal flow of **liquids** across tube bundles, the friction factor is increased if the liquid is being cooled and decreased if the liquid is being heated. The factors previously given for nonisothermal flow of liquids in pipes ("Incompressible Flow in Pipes and Channels") should be used.

For two-phase gas/liquid horizontal cross flow through tube banks, the method of Diehl and Unruh (*Pet. Refiner*, **37**[10], 124–128 [1958]) is available.

Transition Region This region extends roughly over the range $200 < \text{Re} < 2{,}000$. Figure 6-45 taken from Bergelin, Brown, and Doberstein (*Trans. ASME*, **74**, 953–960 [1952]) gives curves for friction factor f_T for five different configurations. Pressure drop for liquid flow is given by

$$\Delta p = \frac{4 f_T N_r \rho V_{max}^2}{2}\left(\frac{\mu_s}{\mu_b}\right)^{0.14} \tag{6-163}$$

where N_r = number of major restrictions encountered in flow through the bank (equal to number of rows when minimum flow area occurs in transverse openings, and to number of rows minus 1 when it occurs in the diagonal openings); ρ = fluid density; V_{max} = velocity through minimum flow area; μ_s = fluid viscosity at tube-surface temperature and μ_b = fluid viscosity at average bulk temperature. This method gives the friction factor within about ±25 percent.

Laminar Region Bergelin, Colburn, and Hull (*Univ. Delaware Eng. Exp. Sta. Bull.*, **2** [1950]) recommend the following equations for pressure drop with laminar flow ($\text{Re}_v < 100$) of liquids across banks of plain tubes with pitch ratios P/D_t of 1.25 and 1.50:

$$\Delta p = \frac{280 N_r}{\text{Re}_v}\left(\frac{D_t}{P}\right)^{1.6}\left(\frac{\mu_s}{\mu_b}\right)^{m}\left(\frac{\rho V_{max}^2}{2}\right) \tag{6-164}$$

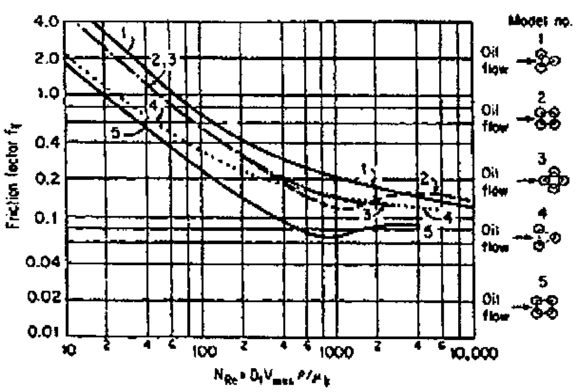
$$m = \frac{0.57}{(\text{Re}_v)^{0.25}} \tag{6-165}$$

where $\text{Re}_v = D_v V_{max} \rho/\mu_b$; D_v = volumetric-hydraulic diameter [(4 × free-bundle volume)/(exposed surface area of tubes)]; P = pitch (= a for in-line arrangements, = a or c [whichever is smaller] for staggered arrangements), and other quantities are as defined following Eq. (6-163). Bergelin, et al. (ibid.) show that pressure drop per row is independent of the number of rows in the bank with laminar flow. The pressure drop is predicted within about ±25 percent.

The validity of extrapolating Eq. (6-164) to pitch ratios larger than 1.50 is unknown. The correlation of Gunter and Shaw (*Trans. ASME*, **67**, 643–660 [1945]) may be used as an approximation in such cases.

For laminar flow of non-Newtonian fluids across tube banks, see Adams and Bell (*Chem. Eng. Prog.*, **64**, *Symp. Ser.*, **82**, 133–145 [1968]).

Flow-induced **tube vibration** occurs at critical fluid velocities through tube banks, and is to be avoided because of the severe damage that can result. Methods to predict and correct vibration problems may be found in Eisinger (*Trans. ASME J. Pressure Vessel Tech.*, **102**, 138–145 [May 1980]) and Chen (*J. Sound Vibration*, **93**, 439–455 [1984]).



Model	Rows	D_o, in	Pitch/D_t
1	10	$\frac{3}{8}$	1.25
2	10	$\frac{3}{8}$	1.25
3	14	$\frac{3}{8}$	1.25
4	10	$\frac{3}{8}$	1.50
5	10	$\frac{3}{8}$	1.50

FIG. 6-45 Friction factors for transition region flow across tube banks. (Pitch is the minimum center-to-center tube spacing.) (*From Bergelin, Brown, and Doberstein*, Trans. ASME, *74, 953 [1952]*.)

BEDS OF SOLIDS

Fixed Beds of Granular Solids Pressure-drop prediction is complicated by the variety of granular materials and of their packing arrangement. For flow of a **single incompressible fluid** through an incompressible bed of granular solids, the pressure drop may be estimated by the correlation given in Fig. 6-46 (Leva, *Chem. Eng.*, **56**[5], 115–117 [1949]), or *Fluidization*, McGraw-Hill, New York, 1959). The modified friction factor and Reynolds number are defined by

$$f_m = \frac{D_p \rho \phi_s^{3-n} \epsilon^3 |\Delta p|}{2 G^2 L (1-\epsilon)^{3-n}} \tag{6-166}$$

$$\text{Re}' = \frac{D_p G}{\mu} \tag{6-167}$$

where $-\Delta p$ = pressure drop
L = depth of bed
D_p = average particle diameter, defined as the diameter of a sphere of the same volume as the particle
ϵ = void fraction
n = exponent given in Fig. 6-46 as a function of Re′
ϕ_s = shape factor defined as the area of sphere of diameter D_p divided by the actual surface area of the particle
G = fluid superficial mass velocity based on the empty chamber cross section
ρ = fluid density
μ = fluid viscosity

As for any incompressible single-phase flow, the equivalent pressure $P = p + \rho g z$ where g = acceleration of gravity z = elevation, may be used in place of p to account for gravitational effects in flows with vertical components.

In creeping flow ($\text{Re}' < 10$),

$$f_m = \frac{100}{\text{Re}'} \tag{6-168}$$

At high Reynolds numbers the friction factor becomes nearly constant, approaching a value of the order of unity for most packed beds.

In terms of S, particle surface area per unit volume of bed,

$$D_p = \frac{6(1-\epsilon)}{\phi_s S} \tag{6-169}$$

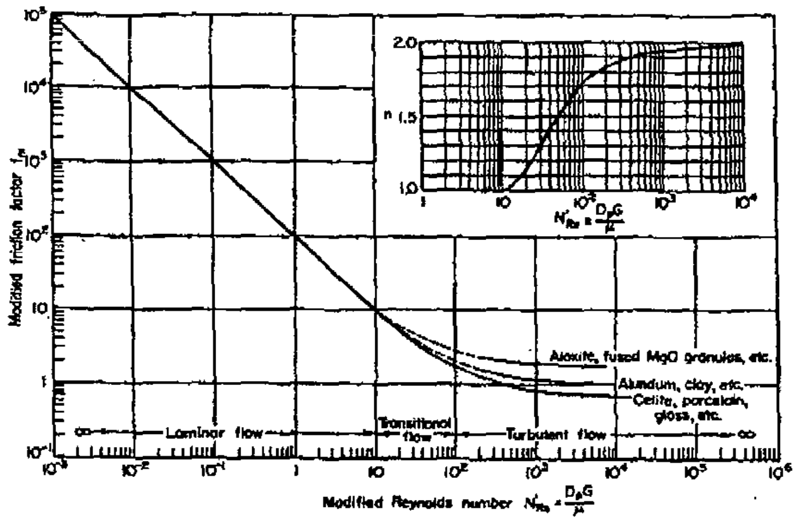


FIG. 6-46 Friction factor for beds of solids. (*From Leva, Fluidization, McGraw-Hill, New York, 1959, p. 49.*)

Porous Media Packed beds of granular solids are one type of the general class referred to as **porous media**, which include geological formations such as petroleum reservoirs and aquifers, manufactured materials such as sintered metals and porous catalysts, burning coal or char particles, and textile fabrics, to name a few. Pressure drop for incompressible flow across a porous medium has the same qualitative behavior as that given by Leva's correlation in the preceding. At low Reynolds numbers, viscous forces dominate and pressure drop is proportional to fluid viscosity and superficial velocity, and at high Reynolds numbers, pressure drop is proportional to fluid density and to the square of superficial velocity.

Creeping flow ($Re' <\sim 1$) through porous media is often described in terms of the **permeability** k and Darcy's Law:

$$\frac{-\Delta P}{L} = \frac{\mu}{k} V \tag{6-170}$$

where V = superficial velocity. The SI units for permeability are m². Creeping flow conditions generally prevail in geological porous media. For **multidimensional flows** through **isotropic** porous media, the superficial velocity **V** and pressure gradient ∇P vectors replace the corresponding one-dimensional variables in Eq. (6-170).

$$\nabla P = -\frac{\mu}{k} \mathbf{V} \tag{6-171}$$

For isotropic homogeneous porous media (uniform permeability and porosity), the pressure for creeping incompressible single phase-flow may be shown to satisfy the LaPlace equation:

$$\nabla^2 P = 0 \tag{6-172}$$

For **anisotropic** or **oriented** porous media, as are frequently found in geological media, permeability varies with direction and a **permeability tensor K**, with nine components K_{ij} giving the velocity component in the i direction due to a pressure gradient in the j direction, may be introduced. For further information, see Slattery (*Momentum, Energy and Mass Transfer in Continua*, Krieger, Huntington, New York, 1981, p. 193–218). See also Dullien (*Chem. Eng. J.* [*Laussanne*], 10, 1–34 [1975]) for a review of pressure-drop methods in single-phase flow. Solutions for Darcy's law for several geometries of interest in petroleum reservoirs and aquifers, for both incompressible and compressible flows, are given in Craft and Hawkins (*Applied Petroleum Reservoir Engineering*, Prentice-Hall, Englewood Cliffs, N.J., 1959). See also Todd (*Groundwater Hydrology*, 2nd ed., Wiley, New York, 1980).

For granular solids of **mixed sizes** the average particle diameter may be calculated as

$$\frac{1}{D_p} = \sum_i \frac{x_i}{D_{p,i}} \tag{6-173}$$

where x_i = weight fraction of particles of size $D_{p,i}$.

For **isothermal compressible flow** of a gas with constant compressibility factor Z through a packed bed of granular solids, an equation similar to Eq. (6-114) for pipe flow may be derived:

$$p_1^2 - p_2^2 = \frac{2ZRG^2T}{M_w}\left[\ln\frac{v_2}{v_1} + \frac{2f_m L(1-\epsilon)^{3-n}}{\phi_s^{3-n}\epsilon^3 D_p}\right] \tag{6-174}$$

where
p_1 = upstream absolute pressure
p_2 = downstream absolute pressure
R = gas constant
T = absolute temperature
M_w = molecular weight
v_1 = upstream specific volume of gas
v_2 = downstream specific volume of gas

For creeping flow of **power law** non-Newtonian fluids, the method of Christopher and Middleton (*Ind. Eng. Chem. Fundam.*, **4**, 422–426 [1965]) may be used:

$$-\Delta p = \frac{150HLV^n(1-\epsilon)^2}{D_p^2\phi_s^2\epsilon^3} \tag{6-175}$$

$$H = \frac{K}{12}\left(9 + \frac{3}{n}\right)^n \left[\frac{D_p^2\phi_s^2\epsilon^4}{(1-\epsilon)^2}\right]^{(1-n)/2} \tag{6-176}$$

where $V = G/\rho$ = superficial velocity, K, n = power law material constants, and all other variables are as defined in Eq. (6-166). This correlation is supported by data from Christopher and Middleton (ibid.), Gregory and Griskey (*AIChE J.*, **13**, 122–125 [1967]), Yu, Wen, and Bailie (*Can. J. Chem. Eng.*, **46**, 149–154 [1968]), Siskovic, Gregory, and Griskey (*AIChE J.*, **17**, 176–187 [1978]), Kemblowski and Merti (*Chem. Eng. Sci.*, **29**, 213–223 [1974]), and Kemblowski and Dziu-